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Critical analysis of techno-economic estimates for the production cost of lignocellulosic bio-ethanol

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ABSTRACT

Bio-ethanol has been claimed to be a green and sustainable alternative to gasoline. The use of food crops on a large scale is ethically unacceptable, but lignocellulosic ethanol has potential to become an alternativfe transportation fuel. This relates to technical issues, and to the eventual cost of bio-ethanol, which requires determination of an absolute production cost. The minimum ethanol selling price (MESP) estimated in different studies varies between \$234 and \$1210 per m³ ethanol (\$0.89 and \$4.58 per gallon), although often the same processing methods are assumed. This entails uncertainties about the potential of bio-ethanol from lignocellulosic sources.

In this study, the main key factors determining these deviations are pinpointed. The assumed values in the different studies were critically investigated and more accurate, unambiguous values proposed. By doing this, a current production cost of \$651 per m³ ethanol was calculated and a realistic projection towards the near future estimates an MESP of \$511. Corn ethanol has already a higher price than the current price of lignocellulosic ethanol due to the high cost of corn. A comparison with gasoline yields a 10% lower price than the future MESP of lignocellulosic ethanol. Due to rising gasoline prices, lignocellulosic ethanol is likely to become in the future not only a more ecological but also a more economical attractive transportation fuel.

Contents

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1. Introduction

Concerns about rapidly increasing oil prices, global warming, depletion of fossil fuels and security of energy supply have stimulated interest in more sustainable energy sources [1]. The combustion of fossil fuels is responsible for more than 70% of the carbon dioxide production [2,3] The transport sector has a major contribution in greenhouse gas (GHG) emissions, of which the impact will continue to increase in the future [4,5]. According to Goldemberg [6], motor vehicles account for 19% of global carbon dioxide (CO_2) emissions [7]. Hence, reducing emissions in this sector would significantly help in reaching targets on climate change. Bio-ethanol, or ethanol derived from biomass, has been recognized as a potential alternative to petroleum based transportation fossil fuels [8]. Furthermore, it is by far the most widely used biofuel for transportation worldwide [9]. Worldwide, countries have become gradually more interested in developing and expanding their biofuel market. As a consequence, the annual world fuel bio-ethanol production has increased remarkably over the last few years from 49 billion (10^9) liters (~13 billion gallons) in 2007 to about 110 billion liters (~29 billion gallons) in 2011 [10].

Even though first generation ethanol has managed to offset some of the gasoline consumption, it has been increasingly criticized. The main reasons are the competition with the food industry and the limited GHG emission savings in comparison to fossil fuels [9,11–13]. Lignocellulosic biomass has been found to be the most promising feedstock for fermentation processes, due to its availability, low cost and the absence of competition with food production [14].

It was found that wordwide, 1623 Tg (10^{12} g) of waste crops and lignocellulosic biomass are potentially available for bio-ethanol production. From these materials, about 491 billion liters of bio-ethanol might be produced, which is about 16 times higher than the current world ethanol production (31 billion liters) [14]. Kim and Dale (2004) studied the global potential bio-ethanol production from lignocellulosic biomass and found that bio-ethanol could replace 353 billion liters of gasoline, which is equivalent to 32% of the global gasoline worldwide consumption, when used in E85 fuel (85% ethanol and 15% gasoline) for a midsized passenger vehicle [2]. Moreover, GHG reductions are projected in the range of 70–85% [15].

Lignocellulosic biomass production is technologically feasible and being tested on a demonstration-scale in many countries including the United States, Spain, Italy, Denmark and Germany [16]. Despite the large amount of research that has been carried out in this field, the economic picture of commercial large-scale lignocellulosic bio-ethanol production still remains uncertain. Therefore many techno-economic models have been developed with the aim of (1) comparing process designs, (2) evaluating the potential of research developments to reduce the production cost, and (3) determining an absolute cost to of lignocellulosic ethanol [17–34]. However, results of these techno-economic models vary significantly from one another, as presented in Fig. 1, where the

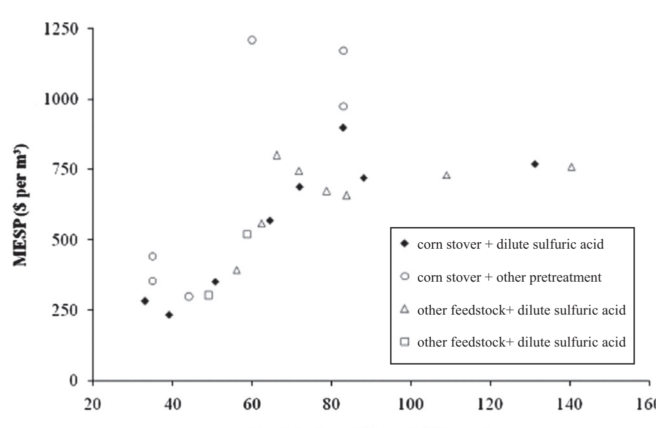


Fig. 1. General comparison of different techno-economic studies: feedstock cost vs. MESP.

minimum ethanol selling price (MESP) is plotted against the feedstock cost, expressed in dollars ($) per dry metric (DM) tonne, of the different studies. Even when the same process technology methods and feedstock are taken into account, the wide scatter in MESP remains.

The aim of this study is to compare the existing techno-economic models and to understand and clarify the causes of this large discrepancy in results. This must result in a better understanding of data obtained from previous studies and will allow making a more accurate estimate of the current ethanol cost and a more realistic projection to the future.

1.1. A side note on units and currencies

Comparing research studies should be easy and straightforward. This is only possible with a consistent use of units, and by preference SI-units, which are an internationally recognized standard [35]. Almost all the techno-economic research studies considered here use non SI-units, such as gallons, short tons or express results per heating value. Furthermore, SI and non-SI units are sometimes mixed. As an example, most research studies express their plant size in metric tonnes (=1000 kg) per day, while feedstock cost and ethanol yield are often expressed per short (U.S.) ton, which is equal to 2000 pounds or 907.2 kg. This of course can easily lead to misinterpreting of reported values. All the reported values in this study, were carefully examined, and are expressed in standard 'base' SI-units or powers of these base units, which will be emphasized through the text. A similar argument holds true for the use of currencies. Most studies express their values in dollar ($), but sometimes values are reported in euro (€) or even local currencies. Since currency conversions daily fluctuate, a simple comparison is difficult. In this work, all values will be expressed in dollar, since most data are provided in this currency.

2. Techno-economic models

In general process models are developed for a selected process scenario in ASPEN Plus™ and process flow diagrams (PFDs) are generated which represent a lignocellulosic refinery. From this, a techno-economic model is derived. A techno-economic model allows to process designs in order to assess the potential of research developments to reduce the production cost. Furthermore, it can also be used for making an estimate of the absolute production cost of lignocellulosic ethanol, based on defined process and plant design assumptions. This estimation can be used to compare lignocellulosic ethanol with other fuels, such as gasoline or first generation biofuels. The overall production process from lignocellulosic biomass to fuel-grade ethanol involves five consecutive steps, i.e., feedstock preparation, pretreatment or first stage hydrolysis, second stage hydrolysis, fermentation and purification [2].

2.1. Processing methods

2.1.1. Feedstock preparation

In the first preparation step, the biomass is cleaned and sized to make its cell structure more accessible for further chemical or biological treatment. This is usually done by a combination of chipping, grinding and milling processes to reach final sizes of 0.2–30 mm depending on the processing methods used in the following steps [36].

2.1.2. Pretreatment or first stage hydrolysis

Lignocellulosic biomass is mainly composed of cellulose, hemicellulose and lignin with smaller fractions of ashes, acids and extractives [7,18,21,25]. In general, cellulose consists of polymeric glucan linkages, while hemicellulose mainly consists of polymeric xylan linkages and smaller amounts of galactan, arabinan and mannan structures. Due to its structure, lignocellulosic biomass requires more aggressive pretreatment methods prior to the actual hydrolysis and fermentation to increase the exposure of cellulose to enzymes during enzymatic hydrolysis [25]. The primary function of this pretreatment step is removing the lignin and hemicellulose structure around the cellulose, rendering it more accessible in following steps [36]. The free hemicellulose in this step is also converted into its soluble sugars, which is mainly xylose. Different pretreatment techniques are applied, such as acid catalyzed hydrolysis, alkaline hydrolysis, steam explosion, liquid hot water (LHW), biological pretreatment or ammonia fiber explosion (AFEX), which are extensively described elsewhere [2,7,18,20,21,25,31]. Currently dilute sulfuric acid is widely regarded as the most economic and efficient pretreatment to commercialize lignocellulosic ethanol in the near future, with xylose yields up to 90% [17,22].

2.1.3. Second stage hydrolysis

In the second stage hydrolysis, the released cellulose of the biomass is converted into glucose. This conversion is typically catalyzed by dilute acid, concentrated acid or cellulase enzymes [2,9]. While acid hydrolysis is a mature technology, enzymatic hydrolysis is projected as the most promising for the near future [21]. During enzymatic hydrolysis, a collection of enzymes called cellulase, is added to the pretreated biomass to hydrolyze the cellulose into hexose sugars. These cellulase enzymes may be either produced in a separate reactor or bought externally from industrial suppliers, such as Novozymes. The high cost of enzyme production and excessive enzymatic dosages are considered to be one of the main challenges for commercialization of a lignocellulosic ethanol industry [37]. Since the research field of enzymatic hydrolysis is still young, many experts see enzymatic hydrolysis as a key factor to produce ethanol in cost-effective way [21]. The conversion of cellulose and hemicellulose can be expressed by the reaction of glucan (for hexoses) and xylan (for pentose) with water:

$$(C_6H_{10}O_5)_n + nH_2O \rightarrow nC_6H_{12}O_6 \quad (1)$$

$$(C_5H_8O_4)_n + nH_2O \rightarrow nC_5H_{10}O_5 \quad (2)$$

The maximum theoretical yield of hexoses and pentose becomes 1.136 kg and 1.111 kg per kg of glucan and xylan, respectively.

2.1.4. Fermentation

Fermentation is the biological process in which the formed hexoses and pentoses are converted into ethanol by a variety of microorganisms, such as bacteria, yeast or fungi. This process occurs under oxygen free conditions and allows microorganisms to obtain energy and to grow. The conversion reaction for hexoses (C6) and pentoses (C5) is as follows:

$$C_6H_{12}O_6 \rightarrow 2C_2H_5OH + 2CO_2 \quad (3)$$

$$3C_6H_{10}O_5 \rightarrow 5C_2H_5OH + 5CO_2 \quad (4)$$

The theoretical maximum yield of both hexoses and pentoses is respectively 0.511 kg ethanol and 0.489 kg CO2 per kg sugar. The overall theoretical ethanol yield (at 20 °C) hence becomes 0.719 and 0.736 l (L) per kg of glucan (or any other 6C structures) and xylan (or any other 5C structures), respectively.

Since most commonly used microorganisms, such as *S. cerevisiae* and recombinant *Z. mobilis*, cannot convert both hexoses and pentoses, genetically engineered recombinant forms are likely to be used in the near-term fermentation process [9,32,38–41].

2.1.5. Purification

The product stream after the fermentation step, denoted as the beer, consists of ethanol, lignin, unconverted hydrocarbons and water, with a typical ethanol concentration in the range of 3–6 wt% [2,18,23,25,32]. The standard purification train involves sending the beer product to a first distillation column (beer column), where the top product consists of 37–40 wt% ethanol and 57–60 wt% water [18,23]. The top product is further purified by a second distillation column to a concentration of ~92.5 wt% and subsequently dehydrated to 99.5 wt% by vapor-phase molecular sieve adsorption [23,42]. One of the main benefits of using distillation to recover bio-ethanol is the high ethanol recovery, i.e.: only 0.1% of the ethanol from fermentation is lost in the bottoms. However, since energy requirements for distillation significantly increase at low ethanol fraction in the beer, alternative technologies, such as liquid/liquid extraction, pervaporation and steam/gas stripping are considered with their pros and cons as described elsewhere [42].

2.1.6. Steam and electricity generation

The bottom product of the first distillation column, which mainly consists of lignin and water, is dried and sent to a boiler for steam and electricity generation. This allows the plant to be self sufficient in energy supply, reduces solid waste disposal cost and generates additional revenue trough sales of excess electricity [18,23,25]. Kazi et al. [25] regarded burning natural gas with a gas boiler for power and heat generation instead of burning the solid residuals with a fluidized bed combustor. The capital investment cost of the gas boiler system was found to be an order of magnitude lower than the fluidized bed combustor. Nevertheless, since natural gas had to be purchased in fairly high amounts, the MESP increased with $48 per m³ compared to burning the solid residuals, even before accounting for disposal costs of the lignin and other plant by-products. This shows that burning the solid residuals for steam and power production is the most beneficial option.

2.1.7. Process integration

In the case of enzymatic hydrolysis, several levels of process integration are possible [21]. The base configuration is defined as Separate Hydrolysis and Fermentation (SHF), in which the liberated cellulose is sent to a different reactor for hydrolysis and subsequent fermentation than the hydrolyzed hemicelluloses and lignin. This configuration has the benefit that each step can be optimized independently and that different microorganisms can be used for fermenting the different sugars. The main disadvantages are the high glucose concentrations, which inhibit fermenting organisms and the cost to construct two separate reactors [17,25]. One way to overcome these drawbacks is using a Simultaneous Saccharification and Fermentation (SSF) configuration, which integrates the cellulose hydrolysis with the fermentation of cellulose. This configuration will reduce the number of reactors and will diminish the inhibiting effect of glucose on fermenting microorganisms. However, a compromise in working temperature has to be chosen since optimum temperatures for enzymatic hydrolysis and fermentation differ [17]. A variation to SHF and SSF is Separate Hydrolysis and Co-Fermentation (SHCF) and Simultaneous Saccharification and Co-Fermentation (SSCF), in which both pentoses and hexoses are fermented by the same microorganism in the same reactor. However, co-fermentation is not yet a mature technology and fermenting microorganisms that produce high ethanol yields from all available sugars are still being developed [21,38–41]. Current ethanol yield of the hemicellulose sugars is rather low, but higher yields are projected in the near future [22]. Hence, SHCF and SSCF are regarded as the most economic configuration to commercialize lignocellulosic ethanol in the short term [18,22,25,26]. While in all previous cases, enzyme production was considered as a separate step, in Consolidated BioProcessing (CBP) all required enzymes and ethanol are produced by a single microorganism in a single reactor. However, currently no such microorganisms or compatible combinations of microorganism exist. Hamlinck et al. [21] predicted it will likely take 20 years of research before CBP will become feasible.

2.2. Cash flow analysis

The absolute production cost of lignocellulosic ethanol is commonly calculated by performing a so called discounted cash flow analysis. Here, the minimum ethanol selling price (MESP) is calculated such that the net present value (NPV) of the lignocellulosic refinery is exactly equal to zero. To perform a cash flow analysis of a lignocellulosic refinery, knowledge of three cost elements are necessary: (1) the total capital investment cost, (2) the variable operating costs, and (3) the fixed operating costs [18,23,25].

2.2.1. Total capital investment

The used methodology to determine the total capital investment is schematically presented in Fig. 2. The capital needed to supply the necessary manufacturing and plant facilities is called the fixed-capital investment (FCI), while that necessary for the operation of the plant is defined as the working capital (WC). The sum of these two cost elements represents the total capital investment (TCI) of the plant [43].

From the generated PFDs, mass and energy balances are developed, which are used to determine the size and specifications of the used process equipments. This allows making an estimation of the equipment cost and installation cost. Equipment costs are commonly obtained from vendor quotes, obtained by the National Renewable Energy Laboratory (NREL) [18,23] This represents the cost of a base case design. Equipment costs of a different size are scaled based on the equipment size for the original price quote, using an appropriate scaling exponent [44]. Subsequently, the installed equipment cost is calculated by applying installation factors, which are often taken from reference books [43,44]. Summation of all equipment results in the total installed equipment cost (TIEC). Once the TIEC is determined, the fixed capital cost can be estimated by taking into account additional costs. These additional costs are typically expressed as a percentage of the TIEC or TCI [18,43]. The working capital depends on the company of interest and most chemical plants use an initial WC in the range of 10–20% of the total capital investment [43]. However, for high turnover products, Garret [44] suggests that using a fraction of the yearly operating cost (equal to the sum of feedstock, fixed and variable cost), typically 10–35% is more relevant.

2.2.2. Variable and fixed operation costs

Analogously, variable costs are also determined by the ASPEN Plus mass and energy balances. An exception to this is the amount of consumed cellulase enzymes, where a direct purchasing cost in $ per m³ ethanol is assumed [22]. In the case of on-site cellulase production, the purchasing cost of enzyme broth is estimated and added to the variable operating cost [25]. Fixed operation costs

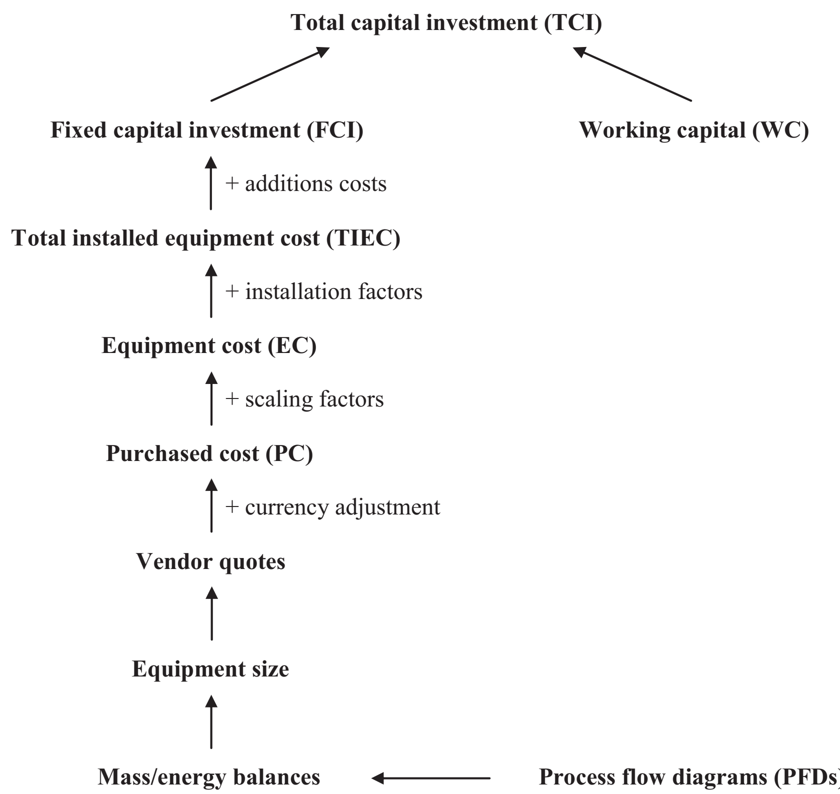


Fig. 2. Schematic representation of the used methodology to determine the TCI.

consist of labor costs, various overhead costs, maintenance costs and insurance costs, which are often estimated as percentages of the total installed equipment cost.

2.2.3. Commercial plant vs. pioneer plant

A commercial plant or n^{th} *commercial plant* is a plant built when technology, plant design and plant operation are well established [45]. Compared with a first kind or *pioneer plant*, the *n*th plant design characteristically minimizes overcapacity and sparing, requires the shortest project schedule to construct and start, and incurs the lowest reasonable engineering cost. The values reported in this, and most other studies, are cost estimates for an *n*th commercial plant. Kazi et al. [25] demonstrated that the absolute ethanol cost of a pioneer plant will be significantly higher than that for an *n*th plant, which is mainly due to a much higher total project investment.

3. Comparison of cost estimations of techno-economic research studies

3.1. Pretreatment step

Although most techno-economic studies assume dilute sulfuric acid as a pretreatment step, other pretreatment methods can be utilized as well. Kazi et al. [25] and Eggeman and Elander [20] both made an economic comparison of the main pretreatment methods, which is presented in Table 1. Both studies assume corn stover as feedstock.

As briefly mentioned in Section 2.1, this analysis confirms that dilute sulfuric acid is the most viable pretreatment option to commercialize lignocellulosic ethanol in the short term. Kazi et al. [25] calculated the investment cost of the liquid hot water method to be much smaller than when dilute sulfuric acid is used (~$6,500,000 vs. ~$36,000,000). This is because of the relative simplicity of the horizontal tubular reactor. However, less hemicelluloses are solubilized, which reduces the efficiency of the enzymatic hydrolysis and therefore results in significantly lower ethanol yields [31]. The two stage acid method has a slightly higher (~4%) investment cost but no cellulase production or purchase costs. Conversion of cellulose to glucose is much lower, which results in a higher projected MESP. The AFEX pretreatment method is only slightly more expensive in comparison to dilute sulfuric acid. This is because the AFEX reactor itself has a lower investment cost, but the additional expense of ammonia recycling results in a higher total investment cost [20,25].

Table 1
Influence of the pretreatment method on MESP.

Reference	Pretreatment method	MESP ($/m³ ethanol)
[20]	Dilute sulfuric acid	354
	LHW	441
	AFEX	372
[25]	Dilute sulfuric acid	898
	Two stage acid	1157
	LHW	1173
	AFEX	975

Also note from Table 1 the large difference in calculated MESP values between the two studies, which is ~3 times higher in the study performed by Kazi et al. [25]. Hence there must be other, more dominant factors that are responsible for the observed discrepancy in MESP values from Fig. 1 and Table 1.

3.2. Feedstock cost

3.2.1. General comparison

From Fig. 1, a general trend of higher MESP with higher feedstock cost is observed. In spite of the relatively low cost of lignocellulosic biomass, the feedstock price remains one of the

major impact factors on the MESP, with estimated contributions between 30% and 40% [25,26]. Hence, an accurate estimate of this value is indispensable to predict the total ethanol cost. The estimated feedstock price will, besides the type of feedstock, mainly depend on the location of the lignocellulosic ethanol plant and the size of the lignocellulosic ethanol plant. The size of the ethanol plant has an impact on the price because average transportation costs increase as the draw radius of the feedstock is higher, which is directly determined by the size and location of the ethanol plant [46]. The feedstock prices used in the different studies are listed in Table 2, together with the type of feedstock.

From Table 2, the differences in feedstock cost are clear, even when the same type (e.g., corn stover) is considered. As can be seen, most studies assume corn stover as feedstock, because of its large abundance. In the United States alone, about 75 million tons of corn stover residues can be harvested annually for the production of bio-fuels [47]. Nevertheless, the assumed corn stover prices of $83 per DM tonne feedstock by Kazi et al. [25] is 2.5 times higher than the value reported by Aden et al. [18].

3.2.2. Cost breakdown

Hess et al. [48] have performed a detailed analysis on the supply chain of corn stover in order to be able to make an accurate estimate of the corn stover price. The economical breakdown of the corn stover cost is presented in Table 3. A detailed description can be found in their work. When all the different cost elements are added an estimated corn stover price of $71.30 per DM tonne feedstock is obtained, which is higher than the value reported in eight out of the ten investigated techno-economic studies. The more recent work of Humbird et al. [23] assumes a feedstock price of $64.50 per DM tonne. This value was obtained from the U.S. Department of Energy [49], which also reported the corn stover price breakdown as presented in Table 3.

Table 2
Feedstock prices used in the techno-economic studies.

Reference	Type of feedstock	Price ($/DM tonne)
[15]	Eucalyptus	62.4
[17]	Corn stover	50.7
[18]	Corn stover	33.1
[19]	Corn stover	66.2
[20]	Corn stover	35.0
[21]—1	Hybrid poplar	58.8
[21]—2	Hybrid poplar	49.0
[21]—3	Hybrid poplar	39.2
[22]—1	Corn stover	72.0
[22]—2	Corn stover	56.1
[23]	Corn stover	64.5
[25]	Corn stover	83.0
[26]	Corn stover	60.0
[30]	Spruce	83.8
[31]	Corn stover	44.1

As can be seen in Table 3, the grower payment and harvesting and collection comprise the two major fractions of the total feedstock cost. The grower payment refers to the financial compensation for the producer for extra fertilizers due to the loss of nutrients from the corn stover. Hence, this grower payment is somewhat variable as fertilizer prices vary considerably from year to year and by region [48]. The harvest and collection cost encompasses all processes related to moving the biomass from the field to the storage or queuing location.

From Table 3, it can also be concluded that the total corn stover prices correspond very well between the different studies, this in contrast to the relative contributions. Deviations between both studies are found in the grower payment, where the SOT price in 2008 and 2010 is found to be 63% higher than the calculated price by Hess et al. [48]. This is because the authors of the latter study used 2006–2008 average fertilizer prices for the calculation of grower payment. They indicate that the grower payment cost would have been $22.30 when 2008 fertilizer prices are used, which is much closer to the values reported by the U.S. Department of Energy [49]. With this adjustment, the overall corn stover price in 2008 would have been $77.7 per DM tonne, which is almost equal to the 2008 SOT value. On the other hand, harvest and collection cost was found to be ~15% and ~30% lower for the 2008 and 2010 SOT case respectively. Nevertheless, no details were provided in their work.

3.2.3. Future cost changes

As can be seen in Table 2, the first scenario of Humbird and Aden [22] assumed a corn stover price of $72.0 per DM tonne feedstock, while the second scenario (4 years later) assumed a 22% lower corn stover price of $56.1. This is because the second scenario estimates the MESP for the near future (2012) and a cost reduction of corn stover was projected. However, no information about the calculation of this cost reduction was provided. Hamelinck et al. [21] also projected a reduction in biomass cost (hybrid poplar) of 16.7% in 5 years and 33.3% in the long term scenario. Again, no details were provided.

Cost reductions seem most likely possible in the harvest and collection step, since the final feedstock cost is most sensitive to the processes performed there [48]. The storage cost of~$8 per DM tonne feedstock, could also be decreased if ethanol refineries would accept multiple feedstocks. This would on the one hand eliminate the need for long-term storage of one seasonal specific feedstock. This would eliminate the need for long-term storage of one seasonal specific feedstock and would dramatically increase the quantity of feedstocks in close proximity to the ethanol plant and reduce the transportation costs [50]. Because of the increasing fertilizer prices it is unlikely that grower payment costs will

Table 3
Economical breakdown of corn stover price (in $ per DM tonne); relative contributions to the total price in parentheses.

Cost element	Cost ($/DM tonne)				
	Reference				
	[46]	**2008 SOT** [47]	**2010 SOT** [47]	**2012 Projected** [47]	**2017 Projected** [47]
Grower payment	15.9 (22%)	25.9 (32%)	25.9 (38%)	25.9 (40%)	33.6 (47%)
Harvesting and collection	21.6 (30%)	18.3 (23%)	15.2 (22%)	14.5 (22%)	14.5 (20%)
Storage	8.1 (11%)	6.9 (9%)	3.9 (6%)	2.7 (4%)	2.7 (4%)
Handling and transportation	11.9 (17%)	16.1 (20%)	12.6 (19%)	12.7 (20%)	12.7 (18%)
Receiving and preprocessing	13.7 (19%)	13.1 (16%)	10.0 (15%)	8.7 (13%)	8.7 (12%)
Total feedstock price	**71.3** (**100%**)	**80.4** (**100%**)	**67.6** (**100%**)	**64.5** (**100%**)	**72.2** (**100%**)

decrease much in the near future. The opposite seems even more likely and the grower payment cost is expected to increase to a value of $33.9 per DM tonne in 2017 [49].

3.3. Overall ethanol yield

As mentioned above, lignocellulosic biomass consists of different polymeric sugar structures. These are first broken down to soluble sugars during the hydrolysis step(s) and are subsequently fermented to ethanol. The overall theoretical ethanol yield (TEY) was calculated to be 0.719 and 0.736 L per kg of 6C structure and 5C structure, respectively.

In practice, no method will achieve a 100% conversion, and hence the actual overall ethanol yield will be always lower than the TEY. Furthermore, during hydrolysis, polymeric structures can be converted to by-products, such as furfural and 5-hydroxymethyl furfural [23]. After the first stage hydrolysis, the hydrolysate slurry is cooled by water and sent to a reactor where either an overliming or ammonia conditioning step is applied. During this step, a small fraction of sugars is lost. Finally, due to contamination, some extra amount of sugars is lost during fermentation.

Not only conversion factors and sugar losses determine the total ethanol yield, but also the biomass composition itself. By comparing different research studies, it is hence wise to look not only at the absolute ethanol yield, but also at the assumed conversion factors and resulting percentage of the theoretical yield that is achieved.

3.3.1. Yield comparison

In Table 4 the reported ethanol yields of the different techno-economic studies together with the obtained MESP are shown. All studies assume dilute sulfuric acid as pretreatment step, with the exception of the second and third scenario of Hamelinck et al. [21] and Sendich et al. [31]. Enzymatic hydrolysis was always assumed for the cellulose to glucose conversion.

From Table 4, a negative correlation ($R = -0.75$) between the overall ethanol yield and MESP was observed. Furthermore, it can be seen that the studies which project values for the future obtain much higher yields than studies that are based on the current years of their studies. Percentages are up to and above 80% for the former and range around 60% for the latter.

3.3.2. Ethanol yield breakdown

Glucan to glucose conversions are assumed to be 90% in most studies [18,21–23,25] with the exception of the first case of Hamelinck et al. [21] where a value of only 80% has been used. Xylan to xylose concentration range from 70% in the study of Dutta et al. [19] up to 90% in the projected cases of Aden et al. [18], Humbird and Aden [22] and Humbird et al. [23]. The conversion of glucose to ethanol is always set higher than or equal to 90%, whereas xylose to ethanol conversion ranges around 85%. Arabinose to ethanol conversion is also set to 85% by Humbird et al. [23], while some other studies project all minor sugars conversion to be as high as the xylose to ethanol conversion [18,21,22,31]. In the CBP scenario of Sendich et al. [31] even all sugar conversions are assumed to be 95%.

Sugar losses can occur during the conditioning step, and depend on the method used. In most designs, an overliming process is used where the pH of the pretreated biomass slurry is increased from 1 to ~10 with lime and then readjusted to ~5 with additional sulfuric acid [23]. Despite the low cost of lime, a relatively high amount of sugars up to 13% can be lost due to side reactions occurring at high pH or pressed out with the formed wet gypsum. The design of Humbird et al. [23] uses a novel ammonia conditioning technique with the aim to avoid these sugar losses [51]. With this technique, the formed hydrolysate can be simply adjusted to the right pH for enzymatic hydrolysis in one step. State of the art technology in 2010 assumed xylose and glucose losses of 2% and 1% respectively, while the target for 2012 was set at 1% and 0%.

A fraction of the fermentable sugars is assumed to be lost during the fermentation step. According to the corn ethanol industry, the tolerated level of loss in a good batch fermentor is 3% and this value is assumed by Aden et al. [18] and Humbird et al. [23]. A more conservative value of 7% loss was used for the modeling calculations by Humbird and Aden [22] and Kazi et al. [25].

Table 4
Absolute ethanol yields, calculated percentages of the theoretical ethanol yield (TEY) and corresponding MESP values of techno-economic studies.

Reference	Year of interest	Ethanol yield (L/DM tonne feedstock)	% of TEY	MESP ($/m³ ethanol)
[15]	2012	340	75.2	560
[17]	2012	375	(79.5)[a]	351
[18]	2010	374	79.5	283
[19]	2009	246	59.6	800
[20]	2012	291	(61.7)[a]	354
[21] – 1	2005–2010	268	58.6	520
[21] – 2	2015–2020	326	71.3	304
[21] – 3	Future	385	84.2	234
[22] – 1	2008	303	64.3	689
[22] – 2	2012	376	79.8	349
[23]	2012	330	76.2	568
[25]	2010	289	64.6	898
[26]	2010	197	–	1210
[31]	2012	290	(61.5)[a]	272
[31]	Future	325	(68.9)[a]	214

[a] Calculated values in parentheses assume the same feedstock composition as Aden et al. [18].

3.4. Enzyme cost

As mentioned earlier, enzymatic hydrolysis is regarded as the most promising cellulose hydrolysis technology. Thus, almost all techno-economic research studies use this process method for their estimation of the MESP. The enzyme cost reflects the cost per gallon ethanol either to produce these cellulase enzymes on-site or to buy them externally from an industrial supplier.

3.4.1. General comparison

In Fig. 3, the MESP is plotted versus the enzyme cost for the techno-economic studies that reported their enzyme cost. Some studies [15,30,31] did not document the enzyme cost in their techno-economic analysis.

From the data presented in Fig. 3, a general trend of higher MESP with higher enzyme cost is observed ($R = 0.71$). The enzyme cost is widely scattered with values ranging from $13.2 for the futuristic case of Hamelinck et al. [21] to $174.4 per m³ ethanol as reported by Kazi et al. [25]. If the enzyme cost is compared to the total MESP, its ratio is 12.8% on average, which stresses the importance of this parameter.

Cellulase enzymes may be either produced in a separate reactor or bought externally from industrial suppliers. The majority of the techno-economic studies assume a fixed cellulose purchasing cost for simplicity, with only a few studies that take on-site cellulase production into account [21,23,25].

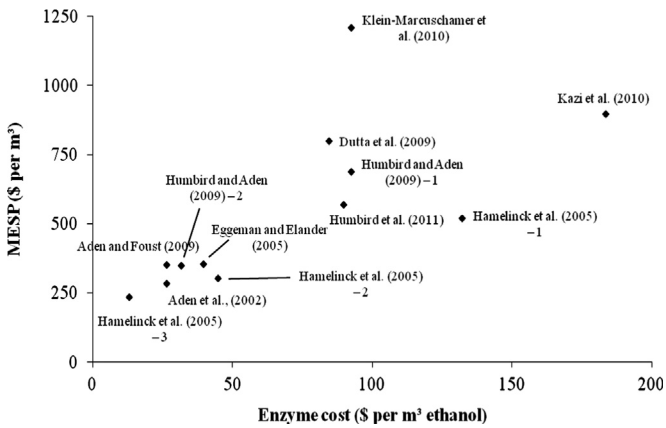


Fig. 3. Comparison of the minimum ethanol selling price versus the enzyme costs reported in techno-economic studies.

3.4.2. *Determination of the cellulase price*

Currently, cellulase is only applied for clothes bleaching, cellulose hydrolysis during coffee bean drying which only need limited amounts of cellulase. Furthermore, these applications are situated in the higher value markets and the small scale production of cellulase makes it therefore an expensive product [21]. Because of this expensive character, some techno-economic studies do not apply current cellulase costs as this would lead to an excessively high enzyme cost, but project enzyme costs for the near future when the large scale production of cellulase has been established.

In 2010, Novozymes, the world's largest enzyme manufacturer, has asserted it can now supply enzymes at a cost of about $132 per m³ ethanol [52]. As can be seen from Table 4, this is higher than most of the reported values. In the work of Humbird and Aden [22], it is even mentioned that the Department of Energy (DOE) and the National Renewable Energy Laboratory (NREL) received feedback from the industry that the purchase price of cellulase enzymes would exceed their assumed value of $92.5 per m³ ethanol. Despite this knowledge, values were not adjusted to more realistic numbers, which can lead to an underestimation of the overall ethanol production cost.

According to the published data of Novozymes, it is clear that many assumed values of enzyme costs are unrealistically low [17,18,20,22]. However, it should be noted that Novozymes did achieve a year-on-year reduction of about 40% in the past few years and claims enzyme costs will continue to decrease significantly in the coming years [53]. Taking into account the current enzyme cost for corn-based ethanol of $8 per m³ ethanol [54], a cellulase purchasing cost lower than $26 per m³ ethanol seems unlikely, since enzymatic hydrolysis of cellulose is known to be more difficult than for starchy-based materials [55].

3.4.3. *On-site cellulase production*

By considering studies that assume on-site enzyme production, Hamelinck et al. (2005) estimated the enzyme cost for the short term scenario (1) by assuming that 5% of the pretreated biomass would be consumed for dedicated on-site cellulase production. This assumption has led to an estimated enzyme cost of $132 per m³ of ethanol. Nevertheless, a separate reactor would have to be constructed, which will increase the capital investment. Kazi et al. [25] compared the scenario of purchasing enzymes to the case with on-site cellulase production, since this could offer economical advantages. In this case 9.2% of the pretreated biomass was consumed and other costs such as the purchase cost of *T. reesei* broth and the capital investment of a dedicated cellulase breeding reactor were also considered. Even with a high enzyme purchasing cost of $184 dollar per m³ ethanol for the base case, the MESP of the on-site cellulase production scenario increased from $898 to $935 per m³ ethanol. This was explained by the high electricity consumption for the supply of air to the enzyme production bioreactor, a 14% increase in the total installed equipment cost and an 11% decreased plant capacity from 200 to 178 million liters per year. Finally, Humbird et al. [23] modeled a cellulase production section, where purchased glucose (corn syrup) is used as primary carbon source for enzyme production. The overall enzyme cost in this configuration was calculated to be $90 per m³ ethanol. Since this is lower than what Novozymes could achieve, they discussed their obtained price with the current enzyme purchase price of $132. The authors concluded that, taking into account extra transport costs, the not sharing of equipment of a standalone facility and higher internal rate of return (IRR) because of the higher risk and lower-volume business, a value of $132 would come out as total enzyme purchasing price. This would mean that a process completely similar to that of Novozymes was modeled, and that they could produce cellulase at the same price. However, this seems very unlikely because the production of cellulase in a single plant will always be less cost-effective in comparison to one of the biggest enzyme manufacturers. Furthermore, the authors even emphasized that an on-site cellulase facility will only make sense in the future and that it seems more likely in the near term that ethanol plants will buy enzymes from an external supplier who has an organization dedicated to improving enzyme performance and reducing costs [21].

3.5. *Fixed capital investment*

As mentioned in Section 2.2.1, a cash flow analysis starts with the determination of the total capital investment, which is the sum of the fixed capital investment and the working capital. In order to compare construction costs from one year to another, the *Chemical Engineering Plant Cost Index* (CEPCI) must be taken into account, since steel and energy prices fluctuate with time [19].

3.5.1. *General comparison*

In Fig. 4, the MESP is plotted against the FCI, adjusted to 2009 US dollars using the CEPCI. All reported studies use dilute sulfuric acid and enzymatic hydrolysis as processing methods. With the exception of Klein-Marcuschamer et al. [26] who assume 2000 metric tonnes per day of wet biomass (15 wt% moisture content), all presented studies assume 2000 DM tonnes of feedstock per day as loading capacity.

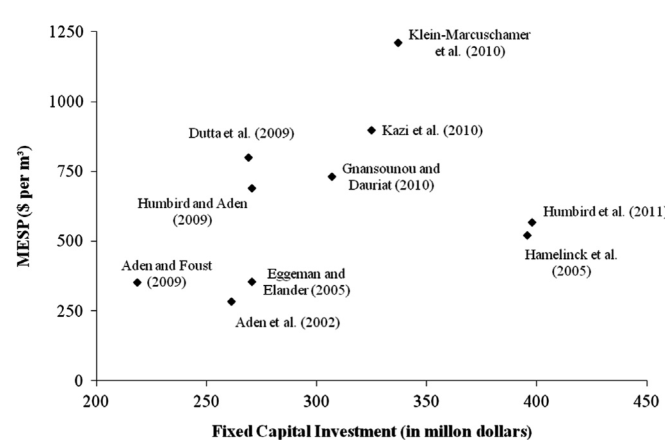


Fig. 4. Comparison of the fixed capital investment versus the minimum ethanol selling price reported in techno-economic studies.

From Fig. 4, a tendency of higher MESP with higher FCI is observed ($R = 0.32$). Despite the adaptation of construction costs to 2009 dollars, the estimated FCI ranges from $218,600,000 in the study of Aden and Foust (2009) to a value of $398,000,000 reported by Humbird et al. [23]. However, when the same processing methods and fixed plant capacity are assumed, a similar refinery would be designed and hence also the investment would be more or less the same. Unfortunately, there are only a limited number of studies that provide detailed information about the calculation of the FCI, which makes comparison of the reported data difficult.

3.5.2. Standard cost calculation methodology

Chem Systems [56] made the first biochemical design for a lignocellulosic ethanol plant, using dilute sulfuric acid and enzymatic hydrolysis as main processing methods. The refinery was furthermore divided into eight sections. The NREL report from Aden et al. [18] used the standard methodology (see Fig. 2) to carefully examine each cost element determining the final FCI. In each section a cost estimate of all necessary equipments (purchased cost) is made. Subsequently, installation factors were obtained from Chem Systems [56] and Harris Group Inc. [57], which determined the total installed equipment cost of the refinery. Finally, additional costs were estimated as proposed by Chem Systems [56] based on industry standards. All determined cost and factors were reviewed by Delta-T corporation and were found to be consistent with other projects the company has completed. Summation of all cost elements resulted in the FCI.

Kazi et al. [25] performed a similar detailed study where different pretreatment scenarios were compared to each other (see Section 3.1), while more recently, Humbird et al. [23] updated the work of Aden et al. [18] and included recent research progress in the conversion areas, optimizations in product recovery, and the latest understanding of the ethanol plant's back end. In contrast to the works of Aden et al. [18] and Kazi et al. [25], an on-site cellulase enzyme section was included in this update to permit better transparency of enzyme economics than the fixed cost contribution assumed in most report designs.

3.5.3. Capital investment cost breakdown

The cost breakdown determining the final TCI of these three detailed studies is compared in Table 5, i.e., a comparison for the cases where dilute sulfuric acid and enzymatic hydrolysis were assumed.

From Table 5, it is clear that the cost elements determining the total direct and indirect cost vary slightly between the study of Aden et al. [18] and Kazi et al. [25]. Hence the sum of total direct and indirect cost only differ by 5%. On the other hand, the purchased equipment cost reported by Humbird et al. [23] is ~1.5 times higher. The reason for this higher value can be attributed to the assumed conditioning step after the pretreatment reactor. Despite the fact that conditioning with ammonia instead of lime results in lower sugar losses, the investment cost of the waste water treatment section increases [23]. In the studies of Aden et al. [18] and Kazi et al. [25], where lime was used as conditioning step, the installed equipment cost (IEC) for wastewater treatment was only ~$4,000,000, while it was no more than $49,000,000 in the report of Humbird et al. [23]. Hence the observed higher purchased cost in the latter studies seems almost completely due to the choice in conditioning step.

Other differences in cost estimation between the studies were found in the hydrolysis and fermentation section. The installed equipment costs of this section were in descending order 31.0, 21.7 and 12.4 million dollars in the studies of Humbird et al. [23], Kazi et al. [25] and Aden et al. [18], respectively. The higher cost of the first study is because a separate hydrolysis and fermentation instead of a simultaneous saccharification and co-fermentation configuration was used, which requires more reactors. If the values of Humbird et al. [23] are compared to the scenario of Kazi et al. [25] where separate C5 and C6 fermentation is considered installed equipment costs differ only slightly. No explanation could be provided for the difference in cost estimation between Aden et al. [18] and Kazi et al. [25]. The IEC of the boiler and turbogenerator section was also higher in the calculation of Humbird et al. [25], but was mainly due to a higher assumed installaon factor, and purchased equipment costs correspond well. Since warehouse and site development costs and indirect costs are calculated as percentages of the TIEC, the trend of total direct and indirect cost is similar, as can be seen from Table 5.

A clear difference between the studies is observed in the values of attributed project contingency and working capital. Aden et al. [18] assumed a contingency factor of 3% of the total direct cost, while 10% was chosen by Humbird et al. [23]. Kazi et al. [25] on the other hand calculated the project contingency as 20% of the total direct and indirect cost, which led to a much higher value. The role of a contingency factor is usually to compensate for possible unforeseen costs or unpredictable events and therefore included in the calculation of the project investment [41]. According to Peters and Timmerhaus [41] contingency factors should range from 5 to 15 percent of the fixed capital investment, with 8 percent being considered a fair average value. In the studies of Aden et al. [18], Humbird et al. [23] and Kazi et al. [25], the assumed values were found to be 2%, 6% and 17% of the FCI, respectively. Hence the lower values can be interpreted as a rather optimistic scenario, while the higher values as more pessimistically oriented.

Finally, the working capital was assumed to be 5% of the FCI in the report of Aden et al. [18] and Humbird et al. [23] while it was 15% in the study of Kazi et al. [25] As mentioned in Section 2.2.1,

Table 5
Comparison of cost elements determining the fixed capital investment and total capital investment (values adjusted to 2009 dollars); relative contributions to the subtotal and total costs in parentheses.

Cost element	Reference		
	[18]	[25]	[23]
Total purchased equipment cost	99,000,000 (62%)	108,700,000 (62%)	153,500,000 (62%)
Total installed equipment cost	**150,500,000 (94%)**	**163,000,000 (94%)**	**230,600,000 (93%)**
Warehouse/site development	10,100,000 (6%)	10,900,000 (6%)	18,200,000 (7%)
Total direct cost	**160,600,000 (63%)**	**174,000,000 (64%)**	**248,800,000 (66%)**
Indirect costs	96,000,000 (37%)	96,700,000 (36%)	126,100,000 (34%)
Total direct and indirect cost	**256,600,000 (98%)**	**270,700,000 (83%)**	**374,900,000 (94%)**
Project contingency	4,800,000 (2%)	54,100,000 (17%)	24,800,000 (6%)
Fixed capital investment	**261,400,000 (95%)**	**324,800,000 (87%)**	**399,700,000 (94%)**
Working capital	13,100,000 (5%)	48,700,000 (13%)	19,900,000 (6%)
Total capital investment	**274,500,000 (100%)**	**373,500,000 (100%)**	**419,700,000 (100%)**

normally a working capital in chemical plants is assumed in the range of 10–20% of the TCI, which is equal to 11–25% of the FCI. However, lignocellulosic ethanol can be considered as a high turnover product, which is expected to be made, shipped and sold in less than 30 days [18]. Hence in the former two studies, the working capital was calculated to be ~22.5% of the yearly operating cost (equal to the sum of feedstock, fixed and variable cost), which corresponded roughly to 5% of the FCI.

3.6. Cash flow analysis parameters

Prior to the start of performing a cash flow analysis some key economic parameters should be defined. In principle, these parameters are the most important specifications of an economic analysis since they form the basis for the calculation of the final MESP. However, most of the assumed values hardly differ from one techno-economic study to another and are based on parameters as specified by Aden et al. [18]. Therefore, these will be briefly outlined first, after which the main differences that require attention are highlighted.

3.6.1. Standard assumed economic parameters

The discount rate is set at 10%, as recommended by Short et al. [58] for renewable energy investments. Tax rate and depreciation method have to be specified to calculate the annual taxes to be paid. Short et al. [58] suggest using the highest federal tax bracket, which is 39% for large corporations in the United States. For the steam and electricity generation unit of the lignocellulosic plant, the 150% declining balance (DB) depreciation method is used for a recovery period of 20 years. The rest of the lignocellulosic refinery makes use of the double declining balance (DDB) method and a recovery period of only 7 years. These depreciation methods are mandated by the Internal Revenue Service (IRS) [18] and are explained elsewhere [59].

The operational plant life is normally set to 20 years and a construction period of 3 years is assumed. Perry and Green [60] indicate that for a moderately complex plant the start-up period should be about 25% of the construction period or 6 months at most. Hence, this maximum value is normally used. During this start-up time, an average of 50% production can be achieved with about 75% expenditure of variable operating expenses and 100% expenditure of the fixed operating expense. The standard design assumes 100% of the financing to come from equity sources and hence nothing is obtained from loans. However, Aden and Foust [17] remark that a more typical business approach would be to fund such a project with a mixture of debt and equity financing.

3.6.2. Parameter differences

Among all investigated studies, the work of Humbird et al. [23] differs the most from these standard assumed parameters. Here, a plant life time of 30 years is used and a slightly lower tax rate of 35%. Furthermore, the start-up period is considered to be three months, in analogy with a mature grain-to-ethanol plant. In their analysis the assumption was also made that the plant would be 40% equity financed based on 8% interest on a 10-year loan. If compared to the original work of Aden et al. [18], it is clear that these modifications are less conservative.

Both studies differ also in the allocation of the costs to the steam plant and general plant for the subsequent depreciation. In order to calculate the general cost for depreciation of the turbo-generator area (area 800), Aden et al. [18] scaled the basic cost with the ratio of the fixed capital investment to the total installed cost. This should be the normal way since additional costs are proportional to the total installed cost and are hence also dependent on the area 800 cost. Humbird et al. [23] only considered the installed cost of area 800 for the depreciation of the steam plant, while all the rest of the FCI is allocated to the general plant. Despite the very detailed work they performed, this is likely due to some inattentiveness. Hence, in the study of Aden et al. [18] the fractional cost of the overall steam plan is equal to ~33.7% of the total plant while this is only ~16.4% in the case of Humbird et al. [23]. Although the effect on the final MESP was calculated to be less than 0.2%, a proper allocation should be performed. To be completely correct, the proper way to determine the allocated cost should be to scale the steam plant with the ratio of the FCI to the total direct cost, since warehouse and site development costs are not proportional to the cost of area 800 [18,23]. In that case, fractional cost of the overall steam plant value would be ~31.6% of the total plant in the study of Aden et al. [18] and ~26.4% in the study of Humbird et al. [23]. In analogy with the explanations provided in Section 3.5, the reason for the lower contribution in the latter study is due to the much higher investment cost of wastewater treatment because of the ammonia conditioning step.

4. Cost estimation of lignocellulosic ethanol

From the analysis above, it is clear that there are different factors responsible for the wide scatter in calculated MESP values from one techno-economic study to another. In this section, first, a more accurate estimate on the current MESP of an *n*th lignocellulosic refinery plant will be made, including a sensitivity analysis discussion. Subsequently, a realistic projection of the future MESP price will be discussed. To conclude, the current and future MESP of lignocellulosic ethanol will be compared to first generation ethanol and traditional gasoline. Prior to starting these calculations, the parameters used will be briefly highlighted. The assumed process design is based on the reports by either Aden et al. [18], Humbird et al. [23] or a combination, because of the detail provided in these works. All the reported values are adjusted to the year 2009 with the appropriate indices [61].

4.1. Determination of process and economic parameters

4.1.1. Current situation

All necessary cost elements for performing a discounted cash flow analysis are presented in Table 6.

Corn stover is considered as feedstock with a feedstock price of $70 per DM tonne. An on-line time of 96% (=8410 h), or slightly more than two weeks of downtime, is considered reasonable for an *n*th plant. Regarding a plant with capacity of 2000 DM tonne corn stover per day, the annual capacity becomes 700,800 DM tonne of corn stover. This corresponds to an annual feedstock cost of $49,100,000.

Table 6
Cost elements used in the discounted cash flow analysis for the current situation; case I: ammonia conditioning, case II: lime conditioning.

Cost element	Case I	Case II
	Annual amount	
Feedstock cost	$49,100,000	$49,100,000
Enzyme purchasing cost	$27,800,000	$26,600,000
Other variable costs	$14,300,000	$13,950,000
Electricity revenues	−$10,400,000	−$14,100,000
Fixed operation costs	$9,850,000	$9,450,000
Total operation cost	**$90,600,00**	**$85,000,000**
	Capital investment	
Total installed equipment	$212,300,000	$171,300,000
Fixed capital investment	$363,700,000	$299,200,000
Total capital investment	**$383,800,000**	**$316,000,000**

As mentioned in Section 3.3, the ethanol yield depends on the individual conversion factors as well as the conditioning step used. The same holds true for the fixed capital investment. Hence two different cases will be investigated, i.e. one with lime as conditioning step (case II) and one with ammonia as conditioning step (case I).

To make yield calculations, the feedstock composition is set fixed at 35% glucan, 20% xylan, 1.5% galactan, 3% arabinan and 0.5% mannan [18,19,23,25,62]. Nevertheless, the whole analysis could be conducted for any given composition. With this composition, the theoretical ethanol yield becomes 435 l per DM tonne feedstock.

Present xylan to xylose conversion and glucan to glucose conversion can be assumed to be equal to 0.85 and 0.9, respectively, according to the SOT of 2010. Sugar losses due to conditioning are set at 2% and 1% for xylose and glucose if ammonia conditioning is considered, and 13% and 12% if overliming is used as conditioning step. Although 3% loss of fermentable sugars was set as a target by Humbird et al. [23], this has not yet been proven. Hence a 5% sugar loss seems more realistic. A conversion of glucose to ethanol of 95% seems currently feasible, as well as 79% for xylose to ethanol. Although arabinose to ethanol conversion is claimed to be 68% in the work of Humbird et al. [23], no evidence for this is provided and research is still ongoing. Hence, the current achievable yield of corn stover, with standard composition as described above, ranges around 287 (lime conditioning—case II) and 300 (ammonia conditioning—case I) liter per DM tonne feedstock. This yield is between 66.1 and 68.8% of the theoretical value.

Cellulase is assumed to be purchased from an enzyme supplier, since this is most likely to be the case in the near term. The purchasing cost is set at $132 per m^3 ethanol produced. Together with an annual ethanol production of 210,500 m^3 and 201,000 m^3 for case I and case II, respectively, the annual cellulase costs become $27,800,000 and $26,600,000. The variable costs for case I were taken from the report of Humbird et al. [23] with subtraction of the cost for on-site enzyme production and are equal to $14,300,000. For case II, variable costs were taken from the work of Aden et al. [18] but adjusted to chemical prices as reported by Humbird et al. [23]. The sum of all variable costs was found to be $13,950,000. Fixed operation costs were determined as described by Humbird et al. [23] and were found to be almost identical in both scenarios.

According to the process flow diagrams of Aden et al. [18], the turbogenerator would produce 30.5 MW of electricity, whereas it was calculated to be 41.3 MW in the report of Humbird et al. [23]. The reason for this is the lower assumed moister content in the stream that is fed to the boiler (44 vs. 52%) and higher assumed boiler efficiency (80% vs. 68%). The plant's energy is calculated to be 23.2 MW and 16.1 MW for case I and case II, respectively, leaving 18.1 MW and 24.6 MW as excess energy. This difference is due to much higher energy consumption for the waste water section in case I, due to the use of ammonia. Taking into account a current electricity price of 6.81 cents per kWh [63], the annual electricity revenues become $10,400,000 and $14,100,000 for case I and case II, respectively.

The installed equipment cost of the different sections is obtained from the study by Humbird et al. [23] with the exception of the pre-treatment section and wastewater section, since these differ between both cases. For case II, the installed equipment cost of these sections was taken from Aden et al. [18]. Additional costs were calculated based on the factors provided by Humbird et al. (2011). This resulted in a total installed equipment cost of $212,300,000 and $171,300,000 for case I and case II, respectively and a corresponding total capital investment of $383,800,000 and $316,000,000.

For the discounted cash flow analysis an operational plant life of 25 years will be assumed, the plant being 50% equity financed based on 8% interest on a 10-year loan. Furthermore, a tax bracket of 37% is chosen and a start-up period of 4.5 months is considered. All other economic parameters were taken from the reference work of Aden et al. [18], since these are used in all other research studies.

4.1.2. Future situation

The current corn stover price was set at $70 per DM tonne. Although some partial costs can decrease in the future, fertilizer prices unlikely will and hence neither is the corn stover likely to decrease significantly. Hence for the determination of the future lignocellulosic ethanol production cost, the feedstock price will be considered unchanged.

Xylan to xylose conversion and glucan to glucose conversion are assumed to go up to a fraction of 0.95 and 0.925, respectively. Glucose to ethanol conversion is assumed to be fixed at the already high value of 0.95 but xylose and arabinose conversions are targeted at 90% [23]. Sugar losses during conditioning are assumed to be reduced by 50% in case of overliming (i.e. 6% glucose, 6.5% xylose and 10% arabinose losses) and set equal to 1% for xylose and arabinose and 0% for glucose [23]. This brings the achievable ethanol yield in five years to 356 and 347 l per DM tonne feedstock for case I and II, respectively.

The enzyme purchasing cost remains the parameter with the highest uncertainty. Although many research studies assume an absolute cost in the range of $26–$32 per m^3 ethanol for the near future [17,18,20,22], this would require an average year-by-year cost improvement of about 25% for five years. Here, an overall reduction in purchasing cost of cellulose of 50% (i.e. $66 per m^3 ethanol) will be assumed in order to not make a too optimistic projection.

As mentioned in Section 3.5.3, it is clear that the purchased equipment cost of a lignocellulosic refinery, if adapted to the same year of interest and compared to a similar process design, hardly differs from one study to another. Hence it is fair to assume that purchased equipments cost, installation costs, additional costs and by consequence also the FCI and TCI will remain constant in the future. The same is assumed for all other costs, such as fixed operating costs, electricity revenues, chemical consumption costs and waste disposal costs. Also the economic parameters from the cash flow analysis for the determination of the current MESP remain unchanged.

4.2. Determination of the absolute ethanol production cost

4.2.1. Current MESP

A discounted cash flow analysis is performed on a lignocellulosic ethanol plant, with specifications as outlined above. It was found that the net present worth of all revenues were exactly equal to the net present worth of all expenses for an MESP of $693 and $651 per m^3 ethanol ($2.62 and $2.46 per gallon) for case I and case II, respectively. The main conclusion that can be drawn from these numbers is that the replacement of lime by ammonia as conditioning step to minimize sugars losses is not justified, since the MESP is $\sim$6.5% higher. This is mainly due to the much higher investment cost and energy requirements of wastewater treatment in case I. If the plant energy demands of case I and II would have been set equal, i.e. less electricity revenues for case II, the MESP of case I would still have been more than 3.5% higher. Although it is claimed by Humbird et al. [23] that ammonia conditioning is a very promising technique, these results prove the contrary. Hence only case II will be further considered for a deeper analysis.

As shown in Fig. 5, the feedstock remains the largest contributor towards the MESP with 38.1%, followed by the cellulase cost (20.6%) and the pretreatment section (18.9%). The boiler and

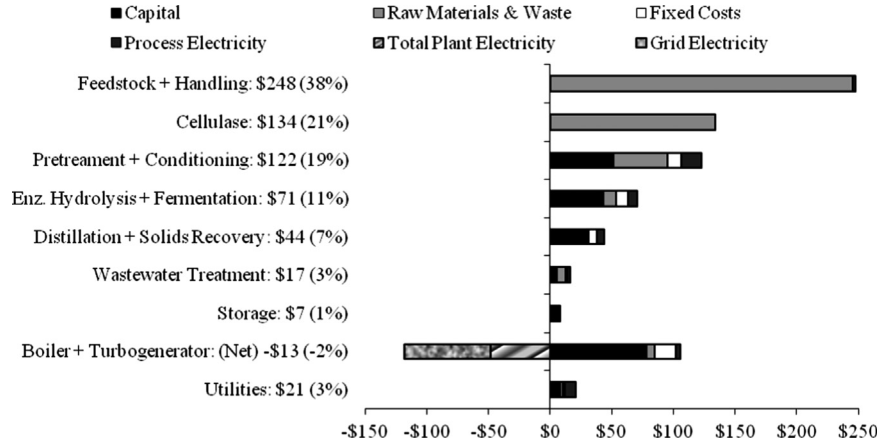


Fig. 5. Overall cost contribution of each process section for the lime conditioning case II (per m³ ethanol); relative contributions to the minimum ethanol selling price in parentheses.

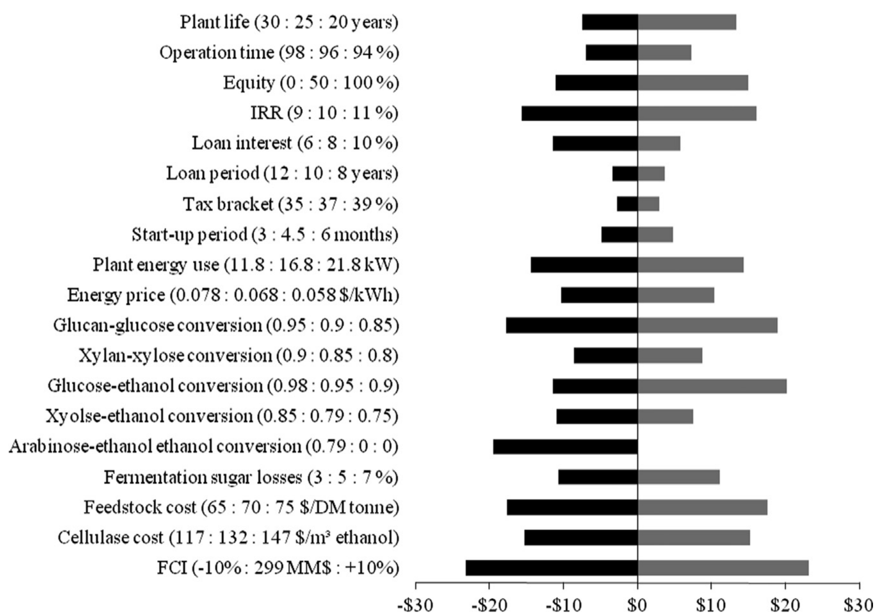


Fig. 6. Influence of variable parameters on the current minimum ethanol selling price for the lime conditioning case II (per m³ ethanol).

turbogenerator accounts also for a large part with 16.2%, but this is more than paid back with electricity production.

Although all elements determining the final MESP were reasoned in detail by analyzing the existing literature, an uncertainty on the exact values always remains. The three detailed studies mentioned in Section 3.5 already performed a so called sensitivity analysis on process-specific parameters which determine the overall ethanol yield and therefore hardly differ qualitatively from one another [18,23,25]. Kazi et al. [25] also investigated the influence of variations in feedstock cost, enzyme purchasing cost, equipment cost, project contingency and electricity price on the MESP. To be complete, besides the most sensitive parameters reported in the former studies, also other parameters will be investigated here, to see whether or not they exhibit an important effect on the MESP. The results of this sensitivity analysis are presented in Fig. 6. In this figure, the middle numbers represent the standard assumed value as outlined in Section 4.1.1, whereas the left and right values are more positive (=less expensive) and negative (=more expensive) variations on the MESP, respectively.

The ethanol yield was found to be most sensitive to the overall cellulose to ethanol conversion (=combination of glucan to glucose conversion and glucose to ethanol), Hence this parameter also had a large impact on the final MESP, i.e. −4.5% and +6% for the given changes. The influence of xylan to ethanol conversion was somewhat smaller, i.e. −3% and +2.5%. Regarding the ethanol yield, it is important to note that if arabinose to ethanol conversion is assumed to be equal to xylose to ethanol conversion (=79%), the MESP would decrease by 3% to $632 per m³ ethanol. Furthermore, the assumed extremes of plant life, operation time, equity, IRR and loan interest are $20.9, $14.3, $26.0, $31.6 and $17.2 apart, respectively. From Fig. 6, the plant energy consumption seems, in economical terms, also a rather unimportant parameter. Even with the relatively wide proposed range of plant energy consumption (10 MW), the influence variation was only around 2%. A similar argument holds true for the electricity prices with variations around 1.6%. Finally, small variations in feedstock and cellulase cost will influence the MESP by ~2.5% and a 10% variation in FCI results in a 3.6% MESP change.

Although it is clear that variations of one parameter have only a limited effect on the MESP, a combination of multiple variable variations might have a much larger effect. Concerning the overall ethanol yield, if all polymeric sugar to monomeric sugar conversions and sugar to ethanol conversion are most pessimistic oriented (i.e., right values in Fig. 2.6), the ethanol yield would

decrease to 258 l per DM tonne feedstock (−10%). This would cause the MESP to increase by ~9% to a value of $710 per m³ ethanol. Analogously, in a very optimistic scenario (i.e., left values in Fig. 2.6), the yield would increase to 327 l per DM tonne feedstock (+14%) and MESP would decrease by 10% to $587 per m³ ethanol.

4.2.2. Future MESP

A similar cash flow analysis is performed in which the only two adjusted parameters are the overall ethanol yield and cellulase purchasing cost as specified in Section 4.1.2.

An MESP of $605.5 per m³ ($2.29 per gallon) ethanol and $495 per m³ ($1.87) per gallon) ethanol was obtained for case I and II, respectively. This is ~12.5% lower than the current estimate for case I while it is ~24% lower for case II. The larger decrease of case II is because in the current scenario a fairly high amount of sugars was lost during the conditioning step and an improvement during this step was projected. Since sugar losses during the conditioning step in case I were already very low, only marginal improvements are possible. This confirms the conclusion drawn in the previous section about the non-promising character of ammonia conditioning. Although the overall yield is higher compared to the overliming process, the much higher investment cost causes the absolute production cost to be always much higher.

In comparison to the current scenario of case II, the relative contributions of the feedstock and pre-treatment step to the MESP have increased from 38% to 41.5% and from 19% to 20.5%, respectively. Because the assumption was made that the cellulase price would decrease by 50% in five years, the cellulase cost contribution has decreased from 21% to 13.5%. As stated above, the cellulase purchasing price is the most uncertain parameter. Hence the influence of its price on the MESP is shown in Fig. 7.

Even at the current cellulase price of $132 per m³ ethanol, the MESP would be $562 per m³ or 13.7% lower than the current price, which accentuates the influence of the overall ethanol yield on the MESP. If the lowest projected purchase price of $26 per m³ ethanol is assumed, the MESP would become $455 per m³, which is more than 30% lower than the current production cost. The absolute difference of $107 per m³ ethanol and relative difference of more than 20% emphasizes the uncertainty associated with the cellulase purchase price.

4.3. Comparison to gasoline and first-generation ethanol

4.3.1. Gasoline production cost

The current retail price of gasoline (January, 2012) is about $3.38 per gallon or $893 per m³ [64]. However, the retail price of gasoline also includes taxes and distribution and marketing costs.

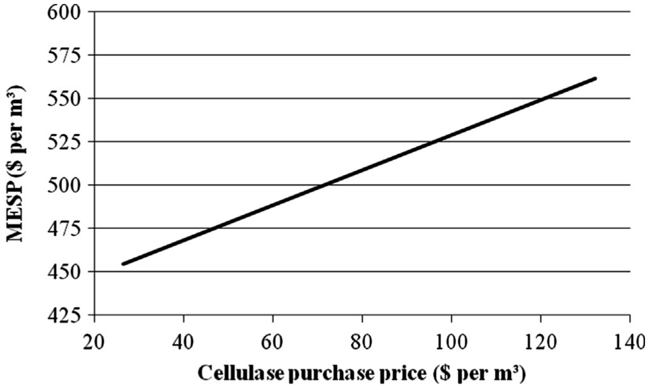


Fig. 7. Influence of cellulase purchase price on the future minimum ethanol selling price for the lime conditioning case II.

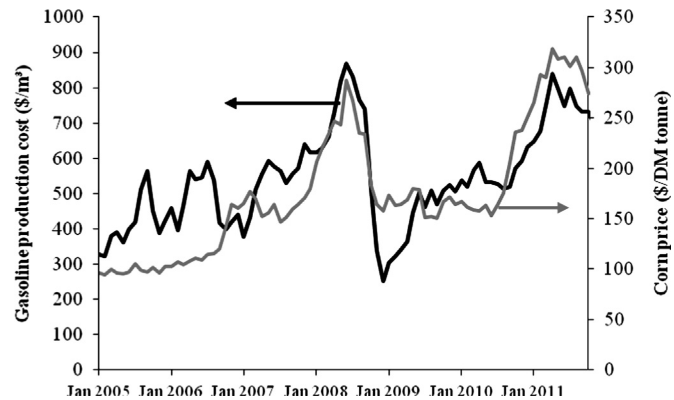


Fig. 8. Monthly average US gasoline production cost and corn price between January 2005 and December 2011 [63].

According to the US Energy Information Administration, in January 2012, taxes and distribution and marketing costs represented on average 12% and 6%, respectively, of the gasoline retail price, but are dependent on the state, the origin of the crude oil and furthermore variable in time [64]. This means that gasoline has currently an effective production cost of about $732 per m³ ($2.77 per gallon). Fig. 8 shows the average US gasoline production cost between January 2005 and December 2011 [65]. From this figure the large fluctuation in price is clear. Only in the year 2011 the boundary costs of gasoline were $621 minimally and up to $880 per m³ maximally. Furthermore, one can remark a sharp increase from January 2009 ($304 per m³) to a value of $798 per m³ in July 2011.

In order to compare gasoline to ethanol, one must take into account the heating values of both fuels, which is ~1.6 higher for gasoline [66]. If this difference is considered, the current gasoline price would be $458 per m³ ($1.73 per gallon) ethanol equivalent. Hence, it is obvious that without any tax credits or subsidies, lignocellulosic ethanol can currently not compete with gasoline on an economic scale since the current MESP ($651 per m³) is more than 40% higher. However, the projected MESP that could be reached in five years ($495 per m³) is only 8% higher than the current gasoline processing price. If the cellulase purchasing prices would decrease more than the assumed 50%, both values will be even closer to each other. Although it is out of the scope of this work to make a projection on the future gasoline price, it is very likely that the future gasoline price will be higher than the current price.

4.3.2. Corn ethanol production cost

As mentioned earlier, several types of feedstock can be used for the production of first generation bio-ethanol [67]. Since corn is the direct counterpart of corn stover, and, furthermore, by far the most abundantly produced bio-fuel in the United States, only corn-based ethanol will be considered here [2,9]. Fig. 8 shows the average US corn prices between January 2005 and December 2011 [65]. In 2011, the average corn price was calculated to be ~$295 per DM tonne.

Hettinga et al. [55] and McAloon et al. [68] have analyzed the production cost of corn ethanol. Despite the details provided in both studies, the corn price at the time of interest ranged around $80–$85 per DM tonne, which is far below the current price. When data of these studies are adjusted to the present corn prices and 2009 dollars, a corn ethanol production cost of $721 per m³ is obtained using the data of McAloon et al. [68], while $865 per m³ was calculated when data of Hettinga et al. [55] were adjusted. In

2007, FO Lichts conducted a worldwide survey of ethanol production cost and found a net production cost in the US of ~$400 per m³ ethanol for corn purchased at $132 per DM tonne [69]. Based on these values, Tao and Aden performed a techno-economic analysis, which confirmed these results [70]. However, if the corn stover price is adjusted to $295 per DM tonne, the total production cost is calculated to be $799 per m³.

These calculations indicate that at current corn prices, the cost of lignocellulosic ethanol is already estimated to be ~18% lower than the total production cost of corn ethanol. Cost contributions between corn stover ethanol and corn ethanol are somewhat different. The feedstock contributes for more than 85% towards the production cost of corn ethanol. Furthermore, the total capital investment of a corn ethanol refinery was found to be more than two times lower than that of a lignocellulosic ethanol refinery. This is due to the fact that the process for making lignocellulosic ethanol is much more complex than for corn ethanol [68].

Also, in a lignocellulosic refinery, a boiler unit is installed for steam and electricity generation from the lignin of the biomass. This accounts for ~35% of the fixed capital investment but allows the plant to be self sufficient in energy supply [18]. Corn ethanol on the other hand has a net energy cost to overcome electricity and heat demand, which accounts for ~54% of the total processing cost [55]. Since this energy is typically obtained by burning fossil fuels, corn ethanol has only limited GHG savings in comparison to gasoline [11].

Finally, from Fig. 8, a correlation ($R=0.79$) between the monthly historical gasoline price and the corn price is observed. This is due to the fact that when gasoline prices increase, more corn is used to produce bio-ethanol. This increase in corn demand will inevitably increase corn and food prices and by consequence leads to even more competition between corn-ethanol and the food market.

4.3.3. Overall comparison

The current production costs of corn stover ethanol, gasoline and corn ethanol are $651, $458, and ~$800 per m³ ethanol, respectively. While gasoline and corn ethanol production are mature technologies, further improvements in the lignocellulose-to-ethanol conversion process are expected. In five years, an absolute cost of $495 per m³ seems achievable, which is only 8% higher than the current gasoline production cost. Dependent on the cellulase purchase prices, the overall cost could become $455 per m³. Furthermore, it is noteworthy that gasoline prices have increased remarkably over the last few years and hence, the future gasoline price is likely to be higher than the current price.

Although corn used to be a much cheaper feedstock in the past, its price has almost tripled over the last decade. As a consequence, corn ethanol has become a relatively expensive product. Since the corn price was found to be related to the gasoline price, a decrease in production cost in the near future seems unlikely. Besides its high production cost, corn ethanol has a net energy cost, which limits the GHG savings and hence its sustainable character.

Lignocellulosic ethanol was found to be 18% cheaper to produce than corn ethanol. Furthermore, it does not compete with the food industry, since lignocellulosic biomass cannot be directly used as a food product [14]. The main advantage of a lignocellulosic ethanol plant lies in the production of excess electricity, which offsets CO_2 emissions from the national power production and has therefore much higher GHG savings compared to corn ethanol [15]. Hence, the large GHG savings of lignocellulosic ethanol and the raising gasoline and correlated corn prices will make lignocellulosic ethanol not only a more sustainable but in the future also an economically more attractive alternative to fossil-based transportation fuels.

5. Conclusions

Lignocellulosic ethanol is a potential transportation biofuel, which does not have to cope with the criticism of traditional first generation biofuels. Many techno-economic research studies have been performed with the aim to minimize the production cost and to determine the final minimal ethanol selling price (MESP). Although often the same processing methods and feedstock are assumed, the forecasted production cost shows a very wide scatter. Mainly the estimated overall ethanol yield, feedstock price and enzyme cost were found to be responsible for the observed differences. Nevertheless, this review has shown that each cost element determining the final ethanol price can be accurately determined and hence also the MESP.

By performing a discounted cash flow analysis, a current lignocellulosic ethanol price of $651 per m³ ethanol was obtained. This value is already lower than the current corn ethanol prices. The projected production cost in five years was found to be slightly higher than the current gasoline price. Furthermore, it is very likely to become lower than the future gasoline production cost, since gasoline prices have increased significantly over the last few years. Additionally, lignocellulosic ethanol offers a very important benefit in comparison to gasoline and corn ethanol, i.e., large GHG savings because of excess electricity production.

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